

# **Early Entrance Coproduction Plant**

## **Phase I**

### **Preliminary Concept Report**

**May 17, 2001**

**DOE Cooperative Agreement**

**No. DE-FC26-99FT40658**

**Book 1**

**Tasks 1-3**



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May 17, 2001

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Attention: Reports Coordination

Subject: DOE Cooperative Agreement No. DE-FC26-99FT40658  
Early Entrance Coproduction Plant  
Phase I – Concept Report  
Letter No. Tx:DOE:020T

Please refer to the subject Cooperative Agreement. As required by Section IV, attachment B, "Federal Assistance Reporting Checklist" attached are two copies of the Phase I Concept Report Sections: Title Page through List of Acronyms which replace the same sections of the concept report. This report is the final report with the placement of these sections into the report submitted on December 28, 2000.

Also enclosed with each copy of the replacement sections, is a CD containing an electronic version of the complete final document in Adobe Acrobat format (pdf).

Should you have any questions, please contact me at 713-752-4754 or by fax at 713-752-4681.

Sincerely,



John H. Anderson  
Acting Project Director

JHA:sms

enclosure

cc: Robert Kornosky – Pittsburgh, PA

# **Early Entrance Coproduction Plant Phase I Concept Report**

Reporting Period: October 1, 1999 – December 31, 2000

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## ABSTRACT

The overall objective of this project is the three-phase development of an Early Entrance Coproduction Plant (EECP) that produces at least one product from at least two of the following three categories:

- Electric power (or heat)
- Fuels
- Chemicals

The objective is to have these products produced by technologies capable of using synthesis gas derived from coal and/or some other carbonaceous feedstock, such as petroleum coke.

The objective of Phase I was to determine the feasibility and define the concept for the EECP located at a specific site and to develop a Research, Development, and Testing (RD&T) Plan for implementation in Phase II. This objective has now been accomplished. A specific site, Motiva Refinery in Port Arthur, Texas, has been selected as the location best suited for the EECP. The accomplishments of Phase I are discussed in detail in this Phase I Concept Report. A RD&T Plan and a preliminary project financing plan have been developed and are submitted separately from this report.

The objective of Phase II is to implement the research, development, and testing work as outlined in the RD&T Plan to enhance the development and commercial acceptance of coproduction technology that produces high-value products, particularly those that are critical to our domestic fuel and power requirements. The project will resolve critical knowledge and technology gaps on the integration of gasification and downstream processing to coproduce some combination of power, fuels, and chemicals from coal and/or other feedstocks.

The objective of Phase III is to develop an engineering design package and a financing plan for an EECP located at the Motiva Refinery in Port Arthur, Texas.

The project's intended result is to provide the necessary technical, economic, and environmental information that will be needed to move the EECP forward to detailed design, construction, and operation by industry.

This report covers all of the subject areas stated in the DOE Cooperative Agreement Statement of Work. It addresses in detail the eight tasks specified and outlines the results, discussions, and conclusions derived from the work performed in each of the areas. The report also recommends areas that we feel need further study in Phase II.

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## EXECUTIVE SUMMARY

This document is the Concept Report for Phase I of the Early Entrance Coproduction Plant (EECP) Project, which is covered by Cooperative Agreement No. DE-FC26-99FT40658 and is part of DOE's "Vision 21" initiative. The objective of this project is the three-phase development of a coproduction plant that produces at least one product from at least two of the following three categories: (1) electric power (or heat), (2) fuels, and (3) chemicals. The objective is to have these products produced by technologies capable of using synthesis gas derived from coal and/or some other carbonaceous feedstock, such as petroleum coke. The objective of Phase I was to determine the feasibility and define the concept for the EECP located at a specific site and to develop a Research, Development, & Testing (RD&T) Plan for implementation in Phase II. Phase I was completed during December 2000. The objective of Phase II is to conduct the research as outlined in the Phase I RD&T Plan to enhance the development and commercial acceptance of coproduction technology that produces high-value products, particularly those that are critical to our domestic fuel and power requirements. The project will help resolve critical knowledge and technology gaps on the integration of gasification and downstream processing to coproduce some combination of power, fuels, and chemicals. Phase II is scheduled for two calendar years (2001 through 2002). Phase III is scheduled for calendar year 2003 and its objective is to develop an engineering design package and a financing plan for the EECP. The overall project's intended result is to provide the necessary technical, economic, and environmental information needed to move the EECP forward to detailed design, construction, and operation by industry.

A consortium, comprised of representatives from Texaco, Inc.; Kellogg Brown & Root, Inc.; General Electric Power Systems; Praxair, Inc.; and Rentech, Inc., was formed to undertake this work. Texaco provided the overall project management as well as conceptual design for the Gasification, Acid Gas Removal, Sulfur Recovery, and the Tail Gas Treating Units of the facility. Kellogg Brown & Root provided engineering/technical services; General Electric provided combined cycle technology; Praxair provided air separation technology; and Rentech provided Fischer-Tropsch synthesis technology. Bechtel, under separate contract with Texaco, provided Fischer-Tropsch product upgrading technology for producing finished wax.

The proposed EECP will coproduce electric power, steam, clean fuels, and finished wax using petroleum coke as the source material. Petroleum coke was selected as the feedstock because it would be the lowest cost source of hydrogen and carbon for the future and the highest probable application of the EECP concept. We believe that the demonstration with petroleum coke will prove the concept, since coal is generally more reactive than coke due to the differences in the molecular structures. The higher volatility of coal, due to the relative ease of its thermal cracking, is perhaps the most obvious difference and is the source of many of the process differences seen. This position is based on gasification pilot plant research and development results for over fifty years of using different feedstocks and their performance in the gasification process. Feedstocks have included petroleum products ranging from natural gas to the heaviest petroleum fractions, petroleum coke, and coal ranging from anthracite to lignite and many types of waste materials. In the EECP concept, approximately 1,235 short tons per day (STPD) of petroleum coke is fed into a Texaco gasifier along with oxygen produced from a Praxair Air Separation Unit (ASU). Inside the gasifier, reactions take place at

very high temperatures, around 1650 K (2500°F), which produce synthesis gas, also known as syngas, a mixture of mainly hydrogen and carbon monoxide, with lesser amounts of water vapor, carbon dioxide, hydrogen sulfide, methane, argon, and nitrogen. The syngas is routed to an Acid Gas Removal (AGR) Unit, where virtually all the sulfur compounds are removed along with some carbon dioxide. Roughly 75% of the cleaned syngas is sent to a General Electric (GE) 6FA gas turbine for power generation. The remainder is routed to a 2.44-meter (8-foot) diameter Fischer-Tropsch (F-T) slurry bubble column reactor (SBCR), where syngas is converted into hydrocarbon liquids. Some unconverted F-T feed gas, along with a light hydrocarbon product produced in the F-T reactor (the combination of which is referred to F-T tail gas), is routed to the gas turbine as fuel for additional power generation. The hydrocarbon liquids are routed to a product upgrading unit to make finished products.

During Phase I, the feasibility of the EECPP concept was evaluated by performing process studies and product slate, marketing, economic, and environmental assessments. Site location criteria were developed, and a specific site was selected. As discussed later in the report, two facilities were considered in some detail for the EECPP site location. The Motiva Port Arthur, Texas petroleum refinery represented a typical refinery application, and the Tampa Electric Company's Polk Power Generation Station near Tampa, Florida, represented a power generation site. For each site, two cases were developed with differing schemes and product slates. The refinery was ultimately selected for further development as the host site. This selection was based on the economic analysis results attained from a model that provided financial return calculations for 17 different scenarios. The economics favored the Port Arthur "Finished Wax" case over the others primarily due to lower feedstock transportation costs, higher product value, and greater infrastructure compatibility. Further process studies, cost estimates, pro forma calculations, and environmental and economic assessments were developed using that site as the basis. The results are summarized in this document. In addition, a Research, Development, and Testing Plan and a Preliminary Project Financing Plan, documented under separate covers, were developed for implementation in Phase II.

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## **EXPERIMENTAL**

There were no experimental methods used in accomplishing the Phase I work. However, in the Phase I write-ups, there are processes discussed that will require further study and probable experimentation during the Phase II work. Some of this experimental work will be accomplished as part of the actions specified in the Research, Development & Testing Plan.



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## Results and Discussion

### Introduction

This section provides an overview of the work performed in Phase I and includes a discussion for the following areas: feedstock considerations, summary of the site selection process, description of the process studies performed, description of the proposed EECF facility, market assessment, environmental assessment, economic assessment, risk assessment, and technical barriers. More detailed information for each of these areas can be found in the appropriate sections for Tasks 2 through 8. Conclusions are discussed in the next section.

### Feedstock Considerations

While the original solicitation requested that coal be used as a feedstock, the analysis of the current available feedstocks resulted in our proposal premise that a petroleum coke feedstock would be the best feedstock to enable the EECF concept to become an actual project. The analysis results were that petroleum coke would be the lowest cost source of hydrogen and carbon for the future. The highest probable application of the EECF concept would be on petroleum coke and most probable at a refinery location due to the high cost of handling and transporting petroleum coke to another location. Therefore, our proposal was that the project would be coal capable and therefore must demonstrate that the design would be capable of converting coal to F-T fuel products. This decision was based on gasification pilot plant research and development results for over fifty years of using different feedstocks and their performance in the gasification process. Feedstocks have included petroleum products ranging from natural gas to the heaviest petroleum fractions, petroleum coke, and coal ranging from anthracite to lignite and many types of waste materials. All of these materials have been gasified successfully. Because of the severe operation conditions used in the gasification process, very high temperature and pressure, it has been determined that there are only minor, in many cases negligible, differences in the reactivities of the various feedstocks.

This universality of performance has been further demonstrated in the more than 130 commercial plants that have been built and run using the Texaco Gasification Process. These plants use the complete range of feedstocks, natural gas, all petroleum fractions, asphalt, petroleum coke, coal, and several waste materials. Any differences in the results, such as variations in the composition of the product syngas or thermal efficiency, can be accounted for by the differences in atomic composition of the feedstocks. Currently, new plants are designed based only on the chemical composition of the feeds.

While this vast store of experience should demonstrate the validity of generalizing gasification performance across feedstocks, some interesting observations have been developed in the cases of petroleum coke and coal. In many other process uses, where operating conditions are less severe, there are significant performance differences. In these cases, coal is generally more reactive than coke because of the differences in the molecular structures. Higher volatility of coal, due to the relative ease of its thermal cracking, is perhaps the most obvious difference and is the source of many of the process differences seen. These processes are generally reaction rate limited at the lower temperatures and pressures used, and the volatiles generated in heating the coal react more rapidly than the solid portions of the material. But in entrained flow

gasification, reaction rates are extremely high, and primarily physical processes, heat, mass transfer and fluid mechanics determine the performance. When these processes are considered, coal and coke are quite similar and hence, they perform essentially the same in the Texaco Gasification Process.

### Summary of Site Selection Process

Shortly after signing of the DOE Cooperative Agreement, efforts to identify potential EECF host sites began. A site selection team was formed to develop specific site criteria. This team was comprised of representatives from Texaco, Kellogg Brown & Root, General Electric, and Praxair. Contacts for potential host sites were also identified and asked to provide liaison with the team. The criteria developed, with relative weighting factors, are shown in Table RD-1.

After preliminary discussions with several possible sites, two were identified as having the potential to match up well with the necessary site criteria. One site, Motiva's Port Arthur, Texas Refinery (PAR) was considered representative of a typical petroleum refinery application. The other, Tampa Electric Company's Polk Power Station near Tampa, Florida, was representative of a power generation facility. The site selection team traveled to these two sites to gain a better understanding of the facilities, available infrastructure, and local conditions. Trips were made to Polk Power Station on December 2, 1999, and to the Port Arthur, Texas Refinery on December 7, 1999. These trips confirmed that both sites were suitable for further consideration as the host EECF site. Each facility satisfied many of the criteria considered important to the EECF concept, such as synergy with existing infrastructure, good construction capabilities and site access, similar environmental requirements, good community relations, etc. Further, representatives from both sites exhibited a strong commitment to the EECF concept and were willing to provide the information necessary to the consortium for the conceptual process engineering studies that were to follow.

**Table RD-1**  
**Site Selection Criteria**

<b>Site Characteristics (5%)</b>
- Size (20%)
- Elevation (10%)
- Geometry (15%)
- Building and Zoning (10%)
- Flood potential (15%)
- Soil data (10%)
- Seismic Zone (10%)
- Climatic conditions (10%)
<b>Transportation Access (5%)</b>
- River Barges (15%)
- Oceangoing Barges (15%)
- Railroad (20%)
- Highway (20%)
- Pipeline (20%)
- Airport (10%)

**Table RD-1**  
**Site Selection Criteria (continued)**

<b>Pollution Control Regulations (8%)</b>
- Air Emissions (20%)
- Liquid Effluents (20%)
- Water Discharge (20%)
- Solid Waste Disposal (20%)
- Hazardous Waste, Toxic Emissions (20%)
<b>Conservation, Community Factors (6%)</b>
- Terrestrial & Aquatic Ecology (20%)
* -- Wetlands
* -- Endangered species
- Obstructions (20%)
* -- Historical Importance
* -- Archeological Importance
-- Corridor Effects
- Local Laws (Noise levels, etc.) (20%)
- Proximity to neighborhoods (20%)
- Proximity to existing reservoir (20%)
<b>Market Factors (28%)</b>
<b>A. Product Values &amp; Marketability (Revenue) (35%)</b>
-- Electricity (25%)
-- Steam (25%)
-- F-T Liquid Fuels (25%)
-- Syngas, Hydrogen (15%)
-- Oxygen (10%)
<b>B. Existing Infrastructure to Supply (35%)</b>
-- Pet coke or coal (60%)
-- Hydrogen (5%)
-- Oxygen (5%)
-- Nitrogen (2%)
-- Treated (demineralized) water (3%)
-- Cooling water (5%)
-- Existing facilities (bldgs, warehouse, control room, lab, maintenance shops, storage) (5%)
-- Existing process facilities (5%)
-- Existing fire protection, medical (5%)
-- Existing utilities availability (5%)
<b>C. Existing Infrastructure to Export (30%)</b>
-- Electricity (40%)
-- Steam (10%)
-- F-T Liquid Fuels (15%)
-- Syngas, Hydrogen (25%)
-- Oxygen (10%)

**Table RD-1**  
**Site Selection Criteria (continued)**

<b>Economic Factors</b>
<b>A. Capital Cost Factors (15%)</b>
- Land (5%)
- Product pipelines (5%)
- Site Preparation (5%)
-- Foundations
-- Grading
-- Drainage & Flood Control
-- Dredging
- Transportation (rail, road, barges) Access (5%)
- Electric Power Connection (5%)
- Water Supply Treatment (5%)
- Liquid Effluent Disposal (5%)
- Site obstructions removal/relocation (10%)
- Materials Handling/Constructability (15%)
- Skilled construction labor availability & cost (40%)
<b>B. Operating Cost Factors (13%)</b>
- Raw materials (feedstock, NG, H <sub>2</sub> ) (50%)
- Utilities (power, cooling water, steam, BFW, condensate) (10%)
- Products distribution (5%)
- Raw Water, catalyst, chemicals, other supplies (5%)
- Labor – operations, maintenance (15%)
- Transportation access maintenance (5%)
- Taxes – property, etc. (5%)
- Tax Incentives (5%)
<b>* Site Commitment to Project (20%)</b>

\* Possible site elimination category.

The criteria are described as follows:

### **Synergy with Existing Infrastructure**

The team gave high importance to the possibility of sharing infrastructure, which included facilities and personnel, with the host facility. Infrastructure was divided into two categories: (1) infrastructure to supply items such as feedstock, hydrogen, oxygen, nitrogen, treated water, cooling water, buildings, warehouse, control room, laboratory, maintenance shops, fire protection, medical facilities, utilities, and (2) infrastructure to export products such as electricity, steam, F-T liquid products, syngas, hydrogen, oxygen, etc. The host site's ability to provide infrastructure was considered vital to the success of the EECF concept.

### **Construction Requirements**

Construction requirements were addressed by the team and included consideration for skilled labor availability and cost, site clearance/preparation, drainage and flood control,

material handling costs, as well as transportation access. These items would be reflected in the overall facility capital cost estimate.

### **Site Access**

The team considered feedstock and product shipping requirements as well as site accessibility for delivery of construction equipment and materials. This included consideration of barge, rail, highway, pipeline, and airport accessibility.

### **Environmental Requirements**

Environmental engineers considered Federal, state, and local environmental regulations for the potential sites. Air emissions, liquid effluents, water discharge, and solid waste disposal were considered. Refer to Section 2.2.2.10 for a description of the environmental considerations.

### **Community**

The team considered the communities associated with the potential sites including the proximity and type of adjacent neighborhoods, zoning requirements, and the host site's relationship with the community. The team also considered the environmental impact on the community in regard to potential wetlands, endangered species issues, historical or archeological significant areas and noise levels.

### **Geotechnic and Topographic Investigation**

Consideration for the site included items such as the space available, geometry, elevation, zoning, flood potential, geotechnical and topographical data, seismic zone, and climatic data.

### **Economic Factors**

Another area considered vital to the success of the concept was capital and operating costs of the EECP. For capital costs, the team made judgements regarding the cost of land, site preparation work, electric power connection, water supply treatment, materials handling, skilled labor availability and cost, and other factors that would ultimately be reflected in total installed cost estimates. Operating costs were also important and consideration was given to feedstock costs, utilities, catalyst/chemicals, product distribution, operating and maintenance labor, taxes, etc.

### **Site commitment to Project**

The host site's commitment to the EECP Project, by both management and operating personnel, was also judged by the site selection team as vital to successful implementation of the concept.

### **Process Engineering Studies for Site Selection**

Two EECP concepts were developed for each of the site, for a total of four cases studied. These cases are detailed further in Task 2 of the Concept Report. Generally, the cases differ mainly in the product slate and the F-TPU technology. In the two Port

Arthur cases, two product upgrading options were developed. In the PARFW case, the F-T liquid is sent to wax finishing unit to produce finished waxes, along with naphtha and diesel. In the PARHCU case, the F-T liquid is sent to a hydrocracking unit, which produces hydrocracked naphtha and diesel. The acid gas generated from the AGR unit is sent to a sulfur plant to make sulfur. In the Port Arthur refinery, steam is exported at different levels.. In the two Tampa cases, two product upgrading options were developed. For the TSC case, the F-T liquid is sent to a Syncrude Dewaxing Unit, where the only hydroprocessing work done is to lower the pour point of the highly waxy F-T liquid. The final product is a single syncrude stream that would be marketed and transported to refineries along the Gulf Coast. For the other Tampa case, THCU, the F-T liquid is sent to a hydrocracker, which produces hydrocracked naphtha and diesel. Also for the Tampa cases, acid gas generated from the AGR Unit is sent to a sulfuric acid ( $H_2SO_4$ ) facility to produce  $H_2SO_4$  for the local fertilizer market. Since there is no market for steam at the Tampa site, all steam is routed to a condensing steam turbine for additional power generation. Table RD-2 summarizes the feedstock, products, and fuels for each of the four cases.

**Table RD-2**  
**Cases Considered**

	<b>PARFW</b>	<b>PARHCU</b>	<b>TSC</b>	<b>THCU</b>
<b>Feed</b>	Petroleum Coke	Petroleum Coke	Petroleum Coke	Petroleum Coke
<b>Standby Fuel – GT</b>	Natural Gas	Natural Gas	Diesel	Diesel
<b>Auxiliary Fuel – HRSG</b>	Natural Gas	Natural Gas & HCU Offgas	Natural Gas	Natural Gas & HCU Offgas
<b>Products</b>	Power Naphtha Diesel Finished Wax Sulfur 6307 kPa steam (900 psig steam) 4238 kPa steam (600 psig steam) 1136 kPa steam (150 psig steam)	Power Naphtha Diesel - Sulfur 6307 kPa steam (900 psig steam) 4238 kPa steam (600 psig steam) 1136 kPa steam (150 psig steam)	Power Syncrude - - Sulfuric Acid - - -	Power Naphtha Diesel - Sulfuric Acid - -

- Case PARFW:** Port Arthur Refinery Finished Wax. The hydrocarbon liquids produced from the F-T reactor are upgraded into three different products – hydrotreated naphtha, diesel, and finished wax.
- Case PARHCU:** Port Arthur Refinery Hydrocracking Unit. The hydrocarbon liquids produced from the F-T reactor are sent to a hydrocracker, which produces diesel and naphtha.
- Case TSC:** Tampa Syncrude. The hydrocarbon liquids produced from the F-T reactor are dewaxed to produce synthetic crude.
- Case THCU:** Tampa Hydrocracking Unit. The hydrocarbon liquids produced from the F-T reactor are sent to a hydrocracker, which produces diesel and naphtha.

Block flow diagrams for each of these cases can be found in Task 2, Figures 2.2.2-1 through 2.2.2-4.

### **Economic Modeling of the Process Cases**

Upon completion of process engineering studies for each site, the final selection was based primarily on financial return calculations produced by an economic model. This model was developed using Microsoft Excel software and provided the ability to change various input parameters and note their affect on the financial calculations. The process studies were conducted first, however, in order to provide the necessary inputs to the model such as product quantities, utility requirements, and capital and operating cost estimates. Prior determination of the site selection criteria helped provide objective supporting information for development of the cost estimates, product market values, transportation costs, inflation rates, etc., that were incorporated in the model.

The inputs used in the economic model included:

- Feedstock prices on an annual basis
- Any product price on an annual basis or an inflator basis
- Capital cost (leveraged and unleveraged)
- Natural gas prices on a yearly basis or by varying inflators
- Maintenance costs on an annual basis to allow for required turnaround periods
- Process availability
- Separate on-stream reliability factors for gasification, F-T Synthesis, and power generation
- Inflation factors for operation and maintenance costs
- Independent electrical inflation rates
- Independent labor rate inflation factors

The model provided the following financial calculations:

- Net Present Value (NPV) – Defined as the current dollar value today of the annual future net cash flows discounted by the cost of capital.
- Present Worth Index (PWI) – Defined as the ratio of the present value of cash inflows to the present value of the cash outflows. PWI measures the relative attractiveness of projects per dollar of investment.
- Present Worth Payout (PWP) – Defined as the time it takes to recover an investment in terms of present value dollars. It represents the elapsed time (expressed in years) it takes for the present value of the net cash inflows to equal the present value of the net cash outflows. It is measured from the initial outflow of funds.
- Discounted Cash Flow Return On Investment (DCFROI) – Defined as the discount rate which equates the project's discounted net cash inflows with its discounted net cash outflows. It can also be interpreted as the return on investment that allows the project's net cash inflows to reduce the present value of the investment to zero.

A total of 17 economic analyses of the refinery and power generation station cases were run with various inputs. Table RD-3 summarizes the results of these calculations.

## Economic Analysis Results of the Process Cases

After the four process cases were developed an economic data were developed for each of the cases. The process data and economic data was inputted into the economic model. Table RD-3 gives a summary of the economic cases and results.

**Table RD-3**  
**Summary of EECF Economic Model Results**

	NPV	PWI	PWP	DCFROI
PARFW Cases:				
• Economic Model Base Case	(49,071,775)	-0.77	N/A	6.17%
• Electric sales @ \$35/MW	(28,868,309)	0.86	N/A	8.04%
• Investment reduction of \$30 MM	(28,785,101)	-0.84	N/A	7.70%
• Operating expense reduction of \$5 MM	(23,749,572)	-0.89	N/A	8.48%
• Investment reduction (\$30 MM), OPEX reduction (\$5 MM), and sales @ \$35/MW	16,740,567	1.09	13.96	12.03%
• Best case with petroleum coke @ -\$10/ton	32,262,309	1.18	12.29	13.41%
• Best case with oil @ \$30/barrel	31,604,405	1.18	12.41	13.33%
• Best case with oil @ \$20/barrel	1,876,730	1.01	15.97	10.67%
PARHCU Economic Model Base Case	(98,286,943)	-0.54	N/A	0.82%
TSC Economic Model Base Case	(158,700,398)	-0.35	N/A	-5.75%
THCU Cases:				
• Economic Model Base Case	(150,949,319)	-0.39	N/A	-4.01%
• Electric sales @ \$35/MW	(114,017,914)	-0.54	N/A	0.64%
• Electric sales @ \$40/MW	(87,677,831)	-0.64	N/A	3.33%
• Electric sales @ \$45/MW	(61,337,748)	-0.75	N/A	5.71%
• Investment reduction (\$30 MM), OPEX reduction (\$5 MM), Electric sales @ \$40/MW	(42,068,955)	-0.80	N/A	6.90%
• Investment reduction (\$30 MM), OPEX reduction (\$5 MM), Electric sales @ \$45/MW	(15,728,871)	-0.93	N/A	9.21%
• Investment reduction (\$45 MM), OPEX reduction (\$5 MM), Electric sales @ \$40/MW	(31,925,618)	-0.84	N/A	7.64%

Overall, the factors having the greatest impact on the selection were:

- The location differentials for petroleum coke was substantial with the petroleum coke having a \$0/ton cost at the refinery location but an \$18/ton cost at the power plant due to transportation and handling costs. Revenues from the sale of electric power were surveyed at the two locations and were higher at the power generation site, but the additional revenue was not sufficient to offset the petroleum coke transportation



cost. In fact, this difference resulted in a substantial economic disadvantage to the EECp concept being located at the power generation site.

- Electric power, steam, and F-T products can be more effectively integrated into a refinery operation than a power generation operation due to greater infrastructure compatibility for the refinery application.
- The typical refinery application has a greater degree of similar type unit operations and the core competencies to incorporate the processes and operations associated with the EECp concept.
- Economies of scale required for a power generation site are larger than the EECp concept design basis. The power generation station application would require a facility at least twice as large as refinery EECp size to have similar economic performance.
- There is an increased thermal efficiency for the EECp concept in a refinery location since the heat from the F-T process can be used directly without any reduction in mechanical efficiency due to conversion to another form of energy.
- Sulfuric acid manufacture required at the power generation site was not economical for the EECp concept design size.

As can be seen, the Port Arthur Refinery "Finished Wax" (PARFW) base case resulted in the best financial performance when compared to the other three base cases for PARHCU, TSC, and THCU. As a result, the PARFW case was considered to have the greatest economic opportunity for deployment of the EECp concept. Subsequently, inputs were varied for this case in order to examine the financial results under different conditions such as; higher rates for electricity sales, reduced capital cost, reduced operating cost, a combination of these assumptions, and various oil prices. The table also shows additional variations run for the THCU case.

Based on these results, the PARFW case at the Port Arthur Refinery was selected for further development for the EECp concept.

### **Description of Process Studies Performed**

In order to develop the EECp concept, it was also necessary to conduct several process studies. Some of these studies were generic, or non-site specific, while others were site-specific. The generic studies included:

- F-T Reactor Feed Gas Optimum Hydrogen ( $H_2$ ) to Carbon monoxide (CO) Ratio
- F-T Product Upgrade Options
- F-T Wastewater Treating/Utilization
- F-T Catalyst/Wax Separation (performed outside the scope of this program)
- F-T Reactor Feed Gas Percent Carbon Dioxide ( $CO_2$ )
- F-T Tail Gas Utilization

Site-specific studies included:

- Process and Heat Integration
- Acid Gas Removal

## The Generic Studies:

### F-T Reactor Feed Gas Optimum Hydrogen (H<sub>2</sub>) to Carbon monoxide (CO) Ratio

The main objective of this study was to determine the optimum F-T feed gas H<sub>2</sub> to CO ratio for the EECP. The F-T Synthesis Unit converts syngas into hydrocarbon liquids using either a cobalt-based catalyst or iron-based catalyst. For the cobalt-based catalyst, the general consensus is that a H<sub>2</sub>:CO ratio of 2 in the syngas feed to F-T is desirable. For the iron-based catalyst, due to water gas shift reaction, there is less certainty of the optimum H<sub>2</sub>:CO ratio. When petroleum coke is fed into the gasifier, the syngas produced from the gasifier has a H<sub>2</sub>:CO ratio of only 0.6 to 0.8. One option is to send the syngas into the F-T Unit without making any adjustment to the H<sub>2</sub>:CO ratio. This option is called the low H<sub>2</sub>:CO ratio feed gas. The other option is to raise the H<sub>2</sub>:CO ratio in the syngas closer to 2, similar to cobalt-based F-T Synthesis and then send it to the F-T Synthesis Unit. This option is called high H<sub>2</sub>:CO ratio feed gas. There are a number of ways to raise the H<sub>2</sub>:CO ratio. The following is a brief description of all the cases studied:

- Case A - Low H<sub>2</sub>:CO ratio: No H<sub>2</sub>:CO adjustment is made to the syngas produced from gasifying petroleum coke. The syngas feed to the F-T reactor has a H<sub>2</sub>:CO ratio of 0.76.
- Case B - High H<sub>2</sub>:CO ratio: Syngas produced from the gasifier is sent to a shift reactor, which shifts a portion of CO present in the syngas into H<sub>2</sub>, thus raising the H<sub>2</sub>:CO ratio to 1.9.
- Case C - High H<sub>2</sub>:CO ratio: Syngas produced from the gasifier is mixed with H<sub>2</sub>-rich syngas produced from a natural gas Steam Methane Reforming Unit (SMR). The combined syngas has a H<sub>2</sub>:CO ratio of 1.9.
- Case D - High H<sub>2</sub>:CO ratio: Syngas produced from the gasifier is mixed with pipeline H<sub>2</sub> to reach a H<sub>2</sub>:CO ratio of 1.9.

To minimize the total number of cases studied, the size of the F-T synthesis reactor, the volumetric flow rate of syngas feed to the F-T reactor, and the gas turbine size were kept the same. The different H<sub>2</sub>:CO ratio in the syngas feed results in different levels of F-T liquid production. The unreacted tail gas was mixed with supplemental syngas from the gasifier. The amount of supplemental syngas was varied for each case to fully load the single GE 6FA turbine. For simplification, only one F-T product upgrading option, hydrocracking, was considered. The F-T liquid produced was sent to a hydrocracker to produce naphtha and diesel. Steam was exported. The power produced from the gas turbine more than meets the internal consumption of the plant and excess power was exported. A -50%/+100% capital cost estimate and operating cost estimates were developed for each case. Economic calculations were then made to determine which case had the highest discounted cash flow return on investment (DCFROI) and net present value (NPV).

For this study, the conclusion was that the lower hydrogen to carbon monoxide ratio (0.76 as compared to 1.9) syngas was more economical for an EECP when the F-T Synthesis is based on use of iron-based catalyst. The economic analysis shows that Case A, the low H<sub>2</sub>:CO ratio case, has highest return on investment and greatest net present value. The expenditures required to increase the hydrogen content of the feed

gas did not improve product yields or quantities sufficiently to improve overall plant economics. Thus, the low  $H_2:CO$  ratio syngas feed, as produced in the gasification section, without any  $H_2:CO$  ratio adjustment was used for the completion of remaining tasks of the Phase I feasibility study. Please refer to Task 2, Attachment C for a summary of the entire study.

### **F-T Product Upgrade Options**

The purpose of this study was to develop flow schemes and cost estimates for upgrading a nominal 2,500 bpsd of F-T liquids. The information from this study was used to define the type of upgrading and corresponding products that would be economically attractive at selected sites, and for subsequent commercial implementation and marketing studies.

Various options for F-T products upgrading were reviewed for their suitability as part of the overall EECF concept. From these options, the following six cases were selected for further study:

- Stabilized Syncrude
- Dewaxed Syncrude
- Hydrocracker (Diesel) product
- Lube Oil Products
- Wax Products
- Alpha Olefins

The overall configurations, the advantages/disadvantages, yields and budget-type capital cost estimates were developed.

The cost of the F-T upgrading plant increases as the complexity of the process increases. The alpha olefin case may be slightly simpler and lower cost than the wax case, but requires more analysis and marketing research. Eventually the site location and the local marketing will determine the preferred type of product upgrading. However, there is an incentive to take advantage of the low-volume, high-value components inherent in the F-T liquids. The high value specialty products are lube oil, wax, and alpha olefins rather than commodity transportation fuels. A niche market for premium "environmental" type diesel is a possible exception that can also be readily implemented.

In general, the F-T liquids can be considered as premium feedstock with exceptional inherent qualities, such as negligible sulfur, nitrogen and aromatic content. Hydroprocessing is directionally the most suitable mode of product upgrading to isomerize (dewax), saturate the olefins and remove oxygenates. With hydrocracking, there is a diesel volume increase of over 10%, due to the wax hydrocracking into lighter components and the addition of  $H_2$ , and a cetane number of more than 70. For this type of diesel there may be a niche market with public transportation companies. Also hydroprocessing for lube oils and/or waxes is suitable for the low volume, high value niche market.

For a complete report of this study please refer to Task 2, Attachment A.

## **F-T Wastewater Treating/Utilization**

The purpose of this study was to develop and evaluate options for treating or utilizing the wastewater stream generated by the F-T synthesis step. The relatively small quantities of mixed alcohol and/or organic acids have little or no market value as chemicals or chemical feedstock.

Three main options, with some variations on some options, for treating and/or utilizing the F-T wastewater stream were evaluated. These included recycling water to the syngas generation unit, pretreatment to remove alcohol before sending the waste stream to the bio-treatment section, and sending the entire waste stream directly to the bio-treatment section.

Cost estimates were not developed for the different options since the recycle option was far superior to other options in eliminating the waste stream by totally recycling it in a manner that involves low capital and operating cost.

Please refer to Task 2, Attachment B, for a complete report of the study.

## **F-T Catalyst/Wax Separation**

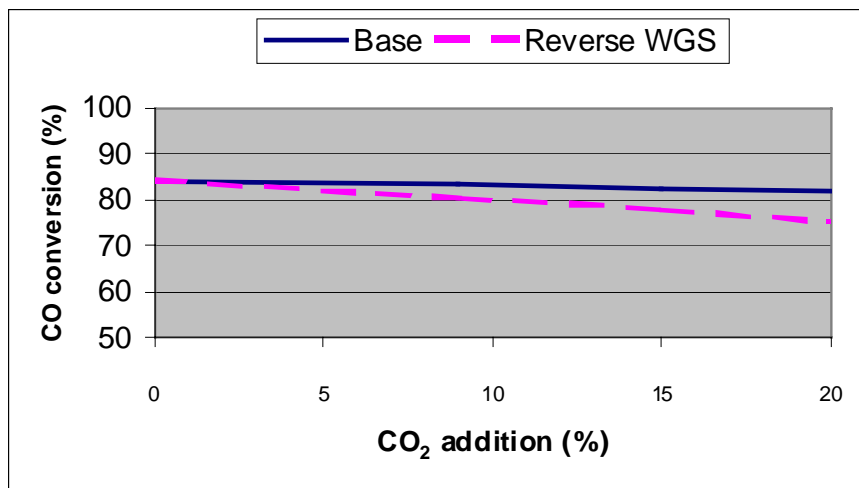
For low temperature F-T reactor operating conditions in a Slurry Bubble Column Reactor (SBCR), the heavier hydrocarbon products (wax) remain in liquid phase, thus increasing the height of the slurry catalyst bed with time. It is, therefore, necessary to remove wax continually to keep the slurry height constant without removing catalyst from the reactor. There is not currently a proven commercial method for removing the catalyst from the wax. This is a critical issue that must be resolved. Currently, various separation methods are being evaluated outside of DOE EECF funding. Prior to the detailed design and construction of the EECF project, sufficient work will have been identified for developing an effective means to separate catalyst and wax. Texaco will demonstrate the effectiveness of the separation on a stand-alone system and a small SBCR outside of DOE funding. Texaco will also privately fund construction and testing of a Demonstration Separator for catalyst/wax separation at the DOE Alternate Fuels Development Unit at La Porte, Texas.

Since the methods planned for catalysis wax separation involve existing proprietary designs developed outside of DOE funding, we will identify the successful method and include the design in Phase III.

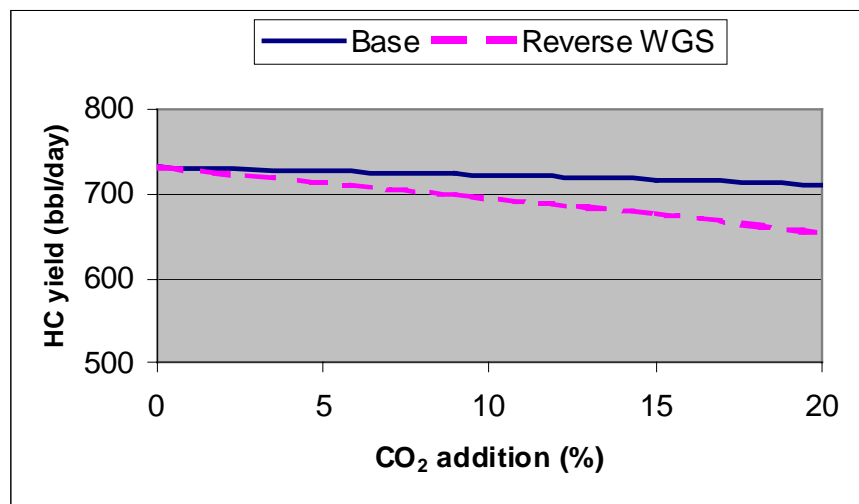
## **F-T Reactor Feed Gas Percent Carbon Oxide (CO<sub>2</sub>)**

The purpose of this study was to evaluate the impact of the presence of carbon dioxide on the performance of the F-T Synthesis Unit and on the EECF. There are currently no experimental data available to evaluate this impact on the Rentech F-T Synthesis Unit. It was proposed to use an in-house numerical model with literature information to estimate the optimal amount of CO<sub>2</sub> in the syngas feed to the F-T Synthesis Unit.

Figures RD-1 and RD-2 show the CO conversion on a mole basis and yield of hydrocarbons.



**Figure RD-1**  
**CO Conversion as a Function of CO<sub>2</sub> in the Feed**



**Figure RD-2**  
**Hydrocarbon Yields as a Function of CO<sub>2</sub> in the Feed**

It is possible that the additional CO<sub>2</sub> induces a reverse water gas shift (WGS) reaction. This reaction will produce more water (relative to syngas), lowering the F-T reaction rate (inhibition by water) and yielding even lower conversions and yields than previously expected. These lower values as a function of the percentage of CO<sub>2</sub> in the feed are depicted as dashed lines in Figures RD-2 and RD-3. The most likely scenario lies somewhere between the two curves. Nevertheless, optimal reactor performance is obtained with no CO<sub>2</sub> in the feed. To validate these results, the effect of CO<sub>2</sub> will be determined in laboratory experiments outside the scope of this work, but its effect or lack of effect will be confirmed and reported in Phase II of the EECF project.

This complete study is included in Task 2, Attachment D.

## **F-T Tail Gas Utilization Study**

The F-T tail gas is comprised of the unconverted syngas, CO<sub>2</sub>, and light hydrocarbons that are produced in the F-T reactor. There are many possible options for F-T tail gas usage. The optimum use will depend on the process scheme and the desired products, which can vary with location. Some possible options for tail gas usage include: recycle to syngas generation, recycle to F-T reactor feed, route as feed to a second F-T reactor, use as fuel for power generation, or use as fuel for steam generation and other fuel requirements.

There are important considerations when addressing tail gas utilization, including:

- The carbon efficiency will be increased by recycling the tail gas allowing for increased conversion and F-T liquid production.
- The process efficiency can be increased by recovering the energy content of the tail gas. Since the tail gas is very clean and free of sulfur and other contaminants this option is environmentally preferred.
- Minimizing emissions to the atmosphere.

Since electric power is a product of the proposed EECF facility, F-T tail gas will be sent to the gas turbine for additional power production.

## **Site Specific Studies:**

### **Process and Heat Integration Study**

The process integration features incorporated at this stage are:

- Recycle F-T wastewater to the Gasification Unit. (Refer to Section 2.1.2.3, "F-T Wastewater Treatment/Utilization.")
- Use of nitrogen from the ASU for the gas turbine (GT) fuel dilution. Surplus nitrogen from the ASU is introduced into the head-end of the gas turbine combustor and injected into the combustion reaction zone. Additional nitrogen required for power augmentation is added downstream into the post-combustion zone.
- Use of nitrogen from the ASU as stripping gas in the AGR. Medium pressure nitrogen from the ASU is used to strip CO<sub>2</sub> from the rich solvent and thereby increase the H<sub>2</sub>S content of the acid gas to the SRU.
- Use of bleed air from the gas turbine as feed to the ASU. The air extraction system provides for the extraction of air from the gas turbine compressor to supply the ASU with compressed air. Air is taken from the outer casing of the combustor after it passes through the combustor liner and transition piece to achieve the required amount of cooling. This results in some additional air pressure drop relative to compressor discharge pressure and also causes the extracted air temperature to be higher than compressor discharge conditions. The machine air extraction is limited to a maximum allowable value, while still providing adequate combustor delta P to drive the cooling flows to the turbine nozzle and combustor components. The air extracted in this manner corresponds to approximately one-quarter of all the air needed for the ASU.

- Utilize oxygen from the ASU in the SRU. Low-pressure oxygen is substituted for air in the SRU to increase the thermal reactor temperature as well as reduce SRU equipment size and cost.

In addition to process stream integration, the following general heat integration steps were taken:

- Approximately 145 MMkJ/hr (138 MMBtu/hr) of low grade heat from the gasifier was used to preheat the deaerator feed water and LP BFW, and to generate 482 kPa (70 psia) steam.
- Steam header levels in the process portion of the EECP were set to match HRSG, steam turbine, and host site header levels to the maximum possible extent.
- Steam usage within the EECP was rationalized to minimize the number of levels at which the steam was exported or imported.

The approach to integration between the EECP and the host site was tailored for the particular site as follows:

### **Port Arthur Refinery**

Preliminary work was done on the combined steam balances of the PAR and the EECP to establish the feasibility of exporting all the surplus steam available from the EECP.

The philosophy was to export all available surplus steam from the EECP (process and HRSG) to the refinery steam header system since the EECP does not have a condensing steam turbine. This resulted in the export of:

- |                                   |                             |
|-----------------------------------|-----------------------------|
| • 140,000 Kg/hr of 6307 kPa steam | (308,000 lb/hr of 900 psig) |
| • 5,400 Kg/hr of 4238 kPa steam   | (12,000 lb/hr of 600 psig)  |
| • 48,500 Kg/hr of 1136 kPa steam  | (107,000 lb/hr of 150 psig) |

A preliminary investigation was made of the steam system at the Port Arthur Refinery (PAR). It consisted of a number of headers fed by a mixture of dedicated fired boilers, waste heat from process units at the refinery, and a number of back-pressure steam turbines, arranged either independently, or in combination with gas turbines. The exhaust from the gas turbine also served as the preheated combustion air stream feeding some of the fired boilers.

It was determined that eliminating a condensing steam turbine from the EECP could decrease capital cost. PAR's extensive steam system can accept steam produced by the EECP. In addition, the refinery can possibly decommission some existing steam generation capacity as a result of importing steam from the EECP. Such decommissioning should reduce the overall operating cost of the refinery.

### **Acid Gas Removal Study**

The purpose of this study was to define the best technologies to remove CO<sub>2</sub>, hydrogen sulfide (H<sub>2</sub>S) and carbononyl sulfide (COS) from the gasification syngas before it enters the F-T Synthesis Unit and the General Electric gas turbine.

Sulfur recovery technologies were selected to match existing technology at each site. A modified Claus SRU with hydrogenation and amine-based tail gas treating was selected for the refinery site and a sulfuric acid plant was selected for the power station site.

Based on past project experience, the following solvent configurations were selected for comparison to be used in the AGR Unit at both sites:

- Case 1 – chemical solvent, methyldiethanolamine (MDEA) at 311 K (100°F)
- Case 2 – same as above except at 297 K (75°F)
- Case 3 – physical solvent, Polyethylene Glycol Dimethyl Ether (PGDE) at 311 K (100°F)
- Case 4 – same as above except at 272 K (30°F)

Cooling the solvent below 311 K (100°F) requires the use of refrigeration. Near identical process configurations of H<sub>2</sub>S absorption, H<sub>2</sub>S concentration, and solvent stripping sections were used for both solvents. This configuration was selected based upon the common performance requirements for high CO<sub>2</sub> diluent recovery for the gas turbine operations, very low-sulfur feed gas for F-T Synthesis, and high H<sub>2</sub>S content in the stripper offgas. The MDEA process requires an upstream COS hydrolysis reactor and a H<sub>2</sub>S scavenger unit to meet the same total sulfur specifications achieved by the process.

Capital costs, initial catalyst and chemical costs, annual utility costs, and annual catalyst and chemical costs were developed for each of the four cases. The capital cost estimates were based on a ratioed, sized equipment list with a projected accuracy of +/- 35%.

For the refinery site, MDEA Case 2 was selected for the following reasons:

- Approximately 2% lower capital cost
- Potential to further reduce capital costs by an additional 2% by combining AGR and Tail Gas Treating Unit (TGTU) amine strippers, although the impact of this on overall reliability needs to be assessed
- MDEA is already in use at the refinery site

For the power station site, MDEA Case 1 was selected for the following reasons:

- Approximately 1-2% lower capital cost
- MDEA is already in use at the power station site

The approximate \$10 MM difference between the sites for overall sulfur removal and recovery is due mostly to the higher cost of the sulfuric acid plant at the power station site.

For site selection, the minimum capital cost MDEA case was suggested as the initial base case.

For details of the complete study, please refer to Task 2, Attachment E.



## Description of Proposed EECF Facility

The selected case to develop the detail study for the Concept Report was the Port Arthur Refinery Finished Wax (**PARFW**) case.

A block flow diagram of the **PARFW** case is shown in Figure RD-3.

The proposed EECF facility will coproduce approximately 55 MW of net electric power for export, approximately 130,000 kg/hr of 4272 kPa (286,588 lb/hr of 620 psia) export steam, approximately 104,000 kg/hr of 1171 kPa (229,209 lb/hr of 170 psia) export steam, approximately 2,025 kg/hr (360 bpd) of finished high-melt wax, approximately 509 kg/hr (97 bpd) of finished low-melt wax, approximately 125 bpd of F-T diesel, approximately 35 bpd of F-T naphtha, approximately 89 short tons per day of sulfur, and consume approximately 1,235 short tons per day of petroleum coke. Additional quantities of power and steam will be produced and consumed internally.

Petroleum coke from the Port Arthur Refinery (PAR) Delayed Coking Unit (DCU) is ground, mixed with water and pumped as a thick slurry to the Gasification Unit. This coke slurry is mixed with high-pressure oxygen from the Air Separation Unit (ASU) and a small quantity of high-pressure steam in a specially designed feed injector mounted on the gasifier. The resulting reactions take place very rapidly to produce synthesis gas, also known as syngas, which is composed primarily of hydrogen, carbon monoxide, water vapor, and carbon dioxide with small amounts of hydrogen sulfide, methane, argon, nitrogen, and carbonyl sulfide. The raw syngas is scrubbed with water to remove solids, cooled, and then forwarded to the Acid Gas Removal Unit (AGR), where the stream is split. One portion of the stream is treated in the AGR to remove CO<sub>2</sub> and H<sub>2</sub>S and then forwarded to the F-T Synthesis Unit. The other portion is treated in the AGR to remove the bulk of H<sub>2</sub>S with minimal CO<sub>2</sub> removal and then forwarded as fuel to the General Electric frame 6FA gas turbine. In the AGR solvent regeneration step, nitrogen from the ASU is used as a stripping agent to release CO<sub>2</sub>. The resulting CO<sub>2</sub> and nitrogen mixture is also sent to the gas turbine, which results in increased power production and reduced nitrogen oxides emissions. The bulk of the nitrogen is also sent to the gas turbine as a separate stream, where its mass flow also helps increase the power production and reduce nitrogen oxide emissions.

Overall, approximately 75% of the sweetened syngas is sent to the gas turbine as fuel. The remaining 25% is first passed through a zinc oxide bed arrangement to remove the remaining traces of sulfur and then forwarded to the Fischer-Tropsch Synthesis Unit. In the F-T reactor, carbon monoxide and hydrogen react, aided by an iron-based catalyst, to form mainly heavy straight-chain hydrocarbons. Since the reactions are highly

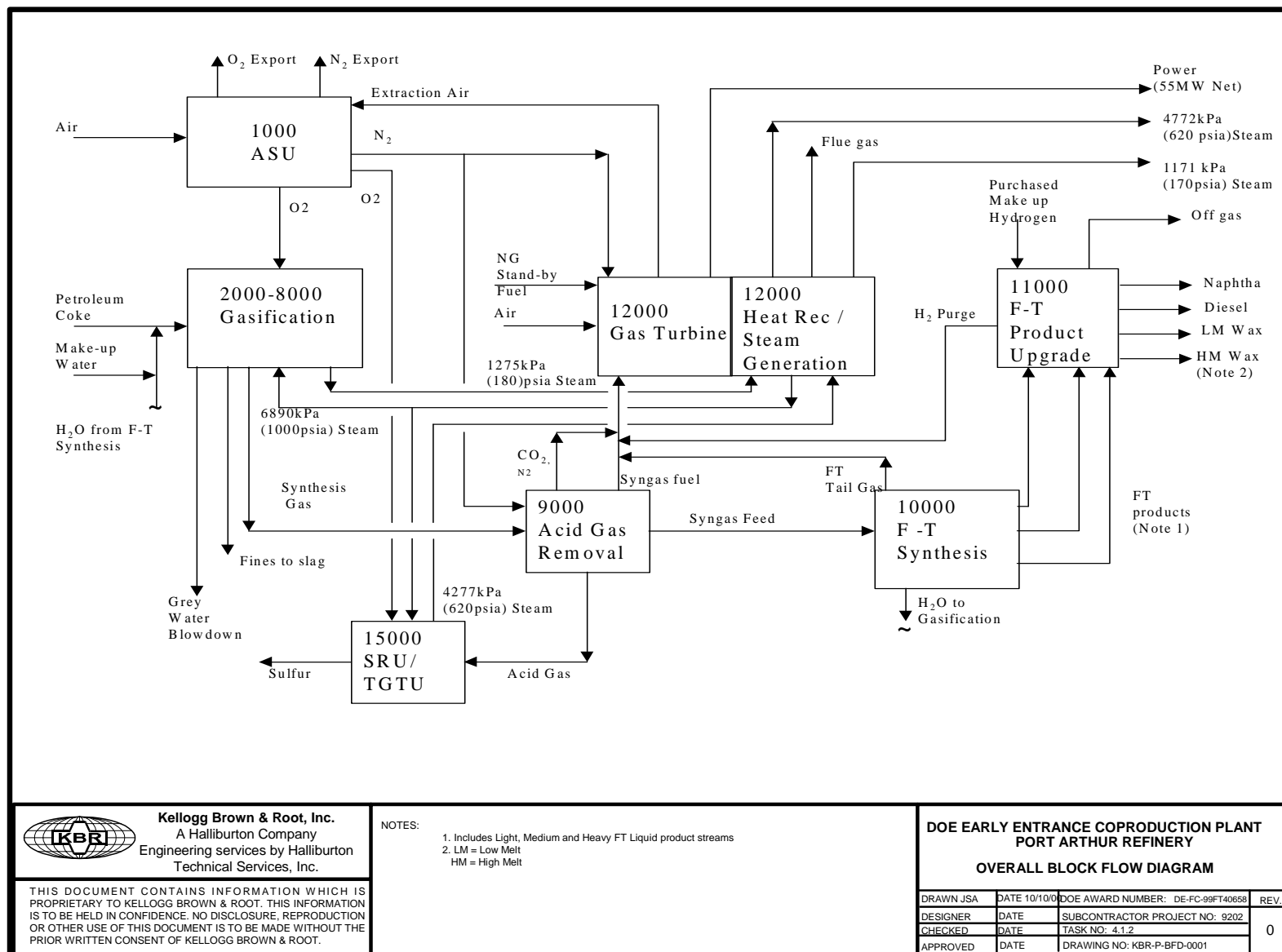


Figure RD-3  
Overall Block Flow Diagram

exothermic, cooling coils are placed inside the reactor to remove the heat released by the reactions. Three hydrocarbon product streams, heavy F-T liquid, medium F-T liquid, and light F-T liquid are sent to the F-T product upgrading unit while F-T water, a reaction byproduct, is returned to the Gasification Unit and injected into the gasifier. The F-T tail gas and AGR offgas are sent to the gas turbine as fuel to increase electrical power production power production by 11%.

In the F-T Product Upgrading Unit (F-TPU), the three F-T liquids are combined and processed as a single feed. The unit consists of a Hy-Finishing™ reactor, product separators, an atmospheric fractionator, naphtha stabilizer, a vacuum distillation tower, feed preheaters, hydrogen compression facilities, and product coolers. The reaction is carried out at elevated pressure and temperature. A mixture of feed hydrocarbons and hydrogen-rich gas is fed to the top of the fixed bed reactor. In the presence of a hydrotreating catalyst, hydrogen reacts slightly exothermally with the feed to produce saturated hydrocarbons, water, and some hydrocracked light ends. The resulting four liquid product streams are naphtha, diesel, low melt wax, and high melt wax. Each of these streams is sent to segregated product storage tanks. The wax tanks are insulated and equipped with steam coils to maintain the storage temperature above the wax melting point. All of the products leave the EECP facility via tank truck. Transfer pumps and truck loading facilities are provided adjacent to the storage tanks.

The power block consists of a GE PG6101 (FA) 60 Hz heavy-duty gas turbine generator and is integrated with a two-pressure level heat recovery steam generator (HRSG) and a non-condensing steam turbine generator. The system is designed to supply a portion of the compressed air feed to the ASU, process steam to the refinery, and electrical power for export and use within the EECP facility. The gas turbine has a dual fuel supply system with natural gas as start-up and backup fuel, and primary fuel as a mixture of syngas from the gasifier, offgas from the AGR Unit, and tail gas from the F-T Synthesis Unit. Nitrogen gas for injection is supplied by the ASU for NO<sub>x</sub> abatement, power augmentation, and the fuel purge system. The combustion system design, including appropriate fuel nozzles, will require new design testing and validation. The heat recovery steam generator is a two-pressure, non-reheat, natural circulation type with horizontal gas flow and vertical fin tubes in all sections. It is arranged with a high-pressure (HP) superheater, HP drum and evaporator, HP economizer, intermediate pressure (IP) superheater, IP drum and evaporator, and IP economizer. High pressure steam is produced at 1,000 psia and intermediate pressure steam at 180 psia.

The Praxair ASU is designed as a single train elevated pressure unit. Its primary duty is to provide oxygen to the gasifier and Sulfur Recovery Unit (SRU), and all of the EECP's requirements for nitrogen and instrument and compressed air. However, it can also export surplus oxygen and nitrogen, when available, to the PAR. As mentioned above, ASU nitrogen product applications within the EECP include its use as a stripping agent in the AGR Unit, as a diluent in the gas turbine where its mass flow helps increase power production and reduce NO<sub>x</sub> emissions, and as an inert gas for purging and inerting. The gas turbine, in return for diluent nitrogen, supplies approximately 25% of the air feed to the ASU, which helps reduce the size of the ASU's air compressor, hence oxygen supply cost.

Acid gases from the AGR, as well as sour water stripper (SWS) offgas from the Gasification Unit, are first routed to knockout drums as they enter the Claus SRU. After entrained liquid is removed in these drums, the acid gas is preheated and fed along with

the SWS gas, oxygen, and air to a burner. In the thermal reactor, the  $\text{H}_2\text{S}$ , a portion of which has been combusted to sulfur dioxide ( $\text{SO}_2$ ), starts to recombine with the  $\text{SO}_2$  to form elemental sulfur. The reaction mixture then passes through a boiler to remove heat while generating steam. The sulfur-laden gas is sent to the first pass of the primary sulfur condenser in which all sulfur is condensed. The gas is next preheated before entering the first catalytic bed in which more  $\text{H}_2\text{S}$  and  $\text{SO}_2$  are converted to sulfur. The sulfur is removed in the second pass of the primary sulfur condenser, and the gas goes through a reheat, catalytic reaction, and condensing stage two more times before leaving the SRU as a tail gas. The molten sulfur from all four condensing stages is sent to the sulfur pit, from which product is transported off site by tank truck.

The tail gas from the SRU is preheated and reacted with hydrogen in a catalytic reactor to convert unreacted  $\text{SO}_2$  back to  $\text{H}_2\text{S}$ . The reactor effluent is cooled while generating steam before entering a quench tower for further cooling. A slip stream of the quench tower bottoms is filtered and sent along with the condensate from the SRU knockout drums to the SWS.  $\text{H}_2\text{S}$  is removed from the quenched tail gas in an absorber by lean methyldiethanolamine (MDEA) solvent from the AGR Unit, and the tail gas from the absorber is thermally oxidized and vented to the atmosphere. The rich MDEA solvent returns to the AGR Unit to be regenerated in the stripper.

The steam system of the EECF serves two main functions: (1) it serves as a medium for exporting a large amount of surplus heat from the EECF to PAR, and (2) it integrates the heat balances between the individual process blocks within the EECF. The system is designed to recover the maximum amount of heat from the EECF and to export it in the form of 4272 kPa (620 psia) and 1171 kPa (170 psia) superheated steam to the refinery. PAR will provide zeolite treated water, which will be passed through a mixed-bed polishing unit to remove final traces of contaminants and produce high-quality deaerator feed water. This water is first preheated by exchange with waste heat streams available from the air separation unit and gasifier, then pumped to the deaerator. Two low-pressure (LP) boiler feed water (BFW) pumps are used to deliver deaerated water to the Gasification Unit where low-grade heat is used to generate 482 kPa (70 psia) saturated steam, which is used within the EECF facility. Other portions of the LP water are sent to the Gasification Unit and HRSG for production of 1275 kPa (185 psia) of saturated steam. A significant portion of this steam is consumed internally within the EECF facility, however, approximately 103,997 kg/hr (229,209 lbs/hr) are superheated in the HRSG and exported to the refinery. The remaining portion of the LP water is passed through the HP boiler feedwater pumps and sent to the HRSG for production of 6890 kPa (1,000 psia) of saturated steam. A small portion of that steam is used in the gasifier and SRU, while the remainder is superheated in a HRSG coil before being let down to 4272 kPa (620 psia) through a steam turbine. The steam turbine is used to drive a generator that recovers 2,827 kW (3,790 horsepower) from the letdown steam. Two IP BFW pumps are used to feed deaerated water to both the 2963 kPa (430 psia) and 4,272 kPa (620 psia) headers after heat exchange with a stream from the gasifier. A portion of the water also picks up heat from the F-T Synthesis Unit to produce 2963 kPa (430 psia) saturated steam for use within the EECF. The remainder is let down to the 1226 kPa (185 psia) header. The remainder of the IP BFW is used to generate 4272 kPa (620 psia) saturated steam, which combines with the letdown from the steam turbine and is exported to the PAR. Provisions are included for letting down steam from one header to the next to facilitate control during start-ups or upset conditions.

A cooling water system is included in the EECF facility. The system consists of a freshwater cooling tower and basin. Motor driven cooling water pumps serve to circulate cooling water between the cooling tower and various cooling water exchangers within the EECF. A vendor-maintained cooling water treatment unit is also provided. Make-up water to the cooling tower is taken from the PAR and cooling tower blowdown is sent to the PAR for treatment.

A flare and associated piping and knockout drum are also provided to handle the full flow of the gasifier in the event that a unit downstream of it should trip.

Firewater and service water systems are provided and connect to the much larger PAR facilities. Services such as sewer, potable water, and oily water systems are also connected to the PAR facilities. In addition, costs for a control room, offices, laboratory, and maintenance shops are included in the EECF capital cost estimate.

The EECF is capable of operation without the F-T Synthesis Unit and the F-TPU. In that case, all of the syngas will be used as fuel to the gas turbine. The gasifier and ASU would then be operating at slightly reduced capacity; however, power and steam export to the refinery would be maintained.

### **EECF Energy Audit**

An energy assessment study using pinch analysis was performed for the EECF facility. The main objective of this study was to carry out an energy audit of the process and identify opportunities to improve energy recovery of the proposed design. The analysis indicated the proposed process configuration exhibited a high degree of recoverable energy efficiency of about 92%. This efficiency value is considered high at this point in design. The detailed results of the study can be found in Task 4, Attachment A.

### **EECF Thermal Efficiency**

On a gross heating value basis, the proposed EECF thermal efficiency has been estimated at 66.5%. This calculation is based on processing 1,235 short tons per day of petroleum coke to produce saleable products: 55 MW of net electrical power; 130,000 kg/hr of 4272 kPa (286,588 lb/hr of 620 psia) steam; 104,000 kg/hr of 1171 kPa (229,209 lb/hr of 170 psia) steam; 2,025 kg/hr (360 bpd) of finished high-melt wax; 509 kg/hr (97 bpd) of finished low-melt wax; 125 bpd of F-T diesel; 35 bpd of F-T naphtha; and 89 tons per day of sulfur.

### **Market Assessment**

Product type, production volumes, market demand, location factors, niche opportunities, and transportation considerations required adequate definition to achieve reasonable estimates of product value and long-term markets. A Product Valuation Team was formed to consider and weigh these factors and to develop methodologies for arriving at realistic current and future price premises for each product at each location.

Products considered for manufacture by the EECF include F-T naphtha, F-T diesel, finished wax, syncrude, power (and steam), and sulfur. For each of these products, the Product Valuation Team considered valuation methodologies, current market size and growth rate, projected new markets, and niche opportunities. The team established a

price basis and forecast for future prices to be used in the project's pro forma calculations. The following is a brief summary of the team's assessment for each product.

### **F-T Diesel**

The current opportunity for F-T diesel is as a blend stock in the distillate market. The U.S. distillate market is a large market of approximately 3.5 MM bbl/day. Of this, about two-thirds of the volume is used in the transportation area in fuels designed to reduce tailpipe emissions. Distillate demand is forecasted to grow approximately 1% or 34,000 bbl/day per year. Of the growth, 88% will occur in the low-sulfur transportation sector. It is expected that U.S. F-T diesel production will displace low-sulfur diesel blend stock imports.

Pricing for F-T diesel, relative to world crude prices, was established by first adding a small premium to diesel prices to account for the low sulfur and cetane benefits. Product prices were projected over a ten-year period based on world crude oil prices published in December 1999 by the EIA Annual Energy Outlook – 2000. Based on a West Texas Intermediate (WTI) price (at Cushing, OK) of \$29.33 per barrel, a blend value of the F-T diesel of \$32.23 per barrel resulted.

In the future, F-T diesel could also be used as a clean fuel alternative in fuel cell automobiles. The absence of sulfur and nitrogen in the fuel eliminates the additional cost and weight of on-board removal components. Chemical applications may also exist for the 9- to 17-carbon normal paraffins. Normal paraffins are primarily used as a building block for linear alkylbenzene sulfonates, a main ingredient in biodegradable detergents. Prices for chemical-grade normal paraffins are typically twice that of high-sulfur diesel. F-T diesel would make a good feedstock for normal paraffin extraction. A narrow cut in the 15- to 17-carbon normal paraffins range would result in a very good synthetic drilling fluid for offshore operations.

### **F-T Naphtha**

F-T naphtha is composed of 21% paraffins, 70% olefins, 9% oxygenates, and is absent of aromatics, naphthenes, sulfur, and nitrogen. There are generally three different uses for naphthas: 1) direct gasoline blending; 2) reforming to high octane reformat for gasoline blending; and 3) steam cracking to ethylene/propylene.

F-T naphtha is expected to have an initial boiling point (IBP) of approximately 319 K (115°F), estimated Reid Vapor Pressure (RVP) of 69 kPa (10 psi), and a research octane number of 53. The high RVP will limit its use as a blending component in summertime gasoline. The low octane number also limits its value as a blending component. F-T naphtha is not considered a suitable reformer feed.

The highest value market for the F-T naphtha will be as a feedstock to the ethylene market and must compete with other feedstocks with current market size estimated at 6 MM tons/year and demand forecasted to grow approximately 5% per year. Based on a WTI crude price of \$29.33, the F-T naphtha value for use as a feedstock to the ethylene market is estimated at \$29.35, which is the price of natural gasoline plus \$0.84 per barrel premium for its high paraffin content.

Possible new markets for the light normal paraffins in F-T naphtha include blowing agents for plastics, solvents in the food industry, drying agents in rubber adhesives, paints, and inks.

### **Finished Wax**

Three finished wax products will potentially be produced from the F-T wax fraction. All are composed of paraffins with a maximum of 0.5% oil content and of food grade quality. The low melt wax melts at approximately 328 K (130°F) and is typically used in candle and paper cup manufacturing. The medium melt wax melts at 333 K to 339 K (140°F to 150°F) and is used in the candle, adhesive, PVC lubricant, and food manufacturing. High melt wax melts in the range 344 K to 350 K (160°F to 170°F) and is used in blending other waxes, adhesives, PVC lubricants, and synthetic rubber. In a refinery, the wax could be used as a high-quality Fluid Catalytic Cracking Unit (FCCU) feed, but at values far below finished wax prices.

The U.S low melt wax market demand is currently 800 MM lb/year or 7,200 bbl/day. The demand is supplied through domestic production and imports. The U.S. medium melt wax market demand is currently 500 MM lb/year or 4,500 bbl/day. The demand is also supplied through domestic production and imports.

The candle industry demand is growing at a rate of 3% per year, led by the high aromatic candle demand. Conversely, the paper cup demand is shrinking 1% per year due to competition from polystyrene cups. Medium melt wax demand is expected to grow at a rate of 2% per year, primarily due to the candle and adhesives markets.

Low melt prices are currently \$0.27 to \$0.29 per pound or \$82 to \$88 per barrel. Medium melt prices are currently \$0.29 to \$0.31 per pound or \$88 to \$97 per barrel. High melt prices are currently \$0.30 to \$0.33 per pound or \$91 to \$100 per barrel. Current high melt prices are depressed due to a short-term oversupply.

Prices for low melt wax are expected to remain at these levels despite increasing demand due to pressure from Far East imports. Imports are also expected to keep medium melt prices relatively flat. High melt prices are expected to recover in approximately two years to \$100 to \$109. After 2005, prices are estimated to increase 1% per year. The lack of imports will help support domestic prices.

### **Syncrude**

Syncrude, derived as a F-T product, was valued using the methodology of comparing to crudes with similar specifications and adjusting for quality and yield differences. The syncrude was compared to light, low-sulfur, United States Gulf Coast (USGC) crude with 36.6 API gravity. A comparison of the theoretical yields, with product values applied, revealed a discount of \$0.40 per barrel for the syncrude. Transportation cost of the syncrude to other USGC locations, in small quantities, further discounted the product another \$3.00 per barrel.

### **Sulfur**

In the U.S. during 1999, approximately 11 million metric tons of elemental sulfur and byproduct sulfuric acid were produced in 149 locations. The value of the shipments was

about \$320 million. World production of all forms of sulfur totals 56 million tons per year. Canada, the second largest sulfur producer in the world behind the U.S. and the world's largest exporter of sulfur, has 11 million metric tons of sulfur stockpiled awaiting more favorable pricing.

Domestic sulfur is produced by the following methods:

- Naturally mined elemental sulfur is known as Frasch sulfur since it is recovered using the Frasch process. Hot water is used to melt the underground element and compressed air is used to transport it to the surface. Only one U.S. company is still producing Frasch sulfur at an offshore mine. Weather conditions and weak market prices have forced many mines to close. Current U.S. Frasch production is less than two million tons per year and is expected to remain relatively unchanged.
- Recovered elemental sulfur is the byproduct of petroleum refineries, natural gas processing plants, and coking operations in over 120 plants. 1999 production was 8 million metric tons.
- Byproduct sulfuric acid is derived from nonferrous metal roasters and smelters. The majority of the roughly one million metric tons/day 1999 production emanates from copper mine smelters.

Worldwide, sulfur supply is very similar to the U.S. market situation. Frasch sulfur production is flat or dropping while recovered and byproduct sulfur supply is increasing due primarily to environmental restrictions that have reduced the overall level of permitted sulfur emissions. Within the U.S., there is a net import and consumption of two million metric tons of elemental sulfur and byproduct sulfuric acid. Eighty percent of the total consumed sulfur is either used as or converted to sulfuric acid and employed as a chemical reagent. Agricultural industries lead in sulfur consumption for phosphoric acid production as an intermediary to phosphatic fertilizers. Petroleum refining is second and copper ore leaching is third in industrial sulfur consumption.

From 1981 to 1998, there was a 0.3 percent growth in the U.S. market. Due to increasing recovered sulfur supply from increasingly strict environmental regulations and weak projected demand for diammonium phosphate fertilizer, a -2% negative growth is projected for the next decade.

From 1981 to 1998, pricing for elemental sulfur has ranged from a high of \$158 per long ton to a low of \$70 per long ton at Tampa, Florida. Current pricing is about \$67.50 per long ton at Tampa. Tampa pricing is normally the highest possible domestic price due to the high demand for sulfur in the Florida area. U.S. West Coast prices could be as low as less than \$1 per long ton. The low price there is due primarily to preparation and shipping costs for overseas delivery. The average value of all domestic shipments was \$29.14 in 1999.

With increased supply primarily from recovered sulfur sold at relatively low prices and slightly decreasing demand due to low growth in the domestic fertilizer industry, a -1% downward price trend is predicted for the next decade.



Sulfur is a major mineral commodity used primarily as a chemical reagent. It has a very well-defined market use in the fertilizer and sulfuric acid industries. Due to its long-term availability, widespread supply and relatively stable current market, no new projected markets of any significant impact are anticipated.

## **Power**

Establishing a current market size for electric power is difficult because of the level of interconnections in the Port Arthur-Beaumont, Texas, area. An assessment of the power companies in the area and the level of interconnections concluded that the market is unlimited for the relatively small size EECF production, with an export level of only approximately 55 MW.

The Port Arthur-Beaumont, Texas area is in the Entergy market arena, interconnecting with The Southern Company, Tennessee Valley Authority (TVA), Southwest Power Pool (SPP), and Southwest Electric Power Company (SWEPCO). In addition, SPP is interconnected to the Mid-Continent Area Power Pool (MAPP). With this level of interconnections, the EECF project should not have any constraints to potential sales as it wheels into any of these market areas.

For this project, power pricing using current dollars was estimated at \$44/MW and descalated at 2.3% per year for four years and limited descalation thereafter.

## **Steam**

Steam pricing is based on the equivalent value of the natural gas required to produce the steam using EIA's delivered natural gas pricing for electrical generation for the low case and a projected \$3.25/Mbtu in 2006 based on current dollars for the high case.

## **Environmental Assessment**

A preliminary assessment of environmental issues associated with the proposed siting of the EECF was undertaken as part of the site selection process in Tasks 2 and 6. The intent of this first assessment was to identify significant differences between the environmental issues at each site so that their impact upon the technical and economic feasibility of an EECF at a site could be properly quantified and factored into the site selection process.

After selection of the Port Arthur Refinery site and more improvement of the integration of the EECF with the host site, the environmental issues were reassessed in Task 7. The work in this task consisted of identifying the relevant environmental requirements that would be imposed on an EECF implementation at Port Arthur, and of quantifying the expected emissions from an EECF. Several opportunities for integration of the environmental control technologies with the host site are identified. This integration results in lower costs without compromising environmental concerns. The overall work accomplished in Task 7 will provide a strong basis for further environmental evaluations as part of Phase III of this project.

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## Economic Assessment

The economics of the EECF were evaluated at two points during Phase I; first during the site selection process (Tasks 2 and 6) and second after the facility configuration was determined for the Port Arthur Refinery site (Task 8). The assessments were performed by using a proprietary economic model that was developed using Microsoft Excel software. It provided the ability to change various input parameters and note their affect on various financial return calculations.

Once the EECF concept was developed for the PARFW case, several runs of the economic model were made for various scenarios. Details of the various runs, including capital and operating cost estimates, along with explanations of their basis, are included in Task 8. The results for the base case show a net present value of \$23.1 MM and an internal rate of return (IRR) of 11.9%. The evaluation from preliminary financing review indicates that an IRR of 15% to 15.5% would be required for this project to obtain consideration.

A potential improvement in economics involves the role Government and/or Government incentives could play. DOE's assistance in supporting incentives or tax relief could be beneficial. These incentives could include:

- DOE's support and assistance in proposing Federal legislation to exempt F-T fuels and blends from Section 4081 Federal Motor Fuel Excise Tax, and
- DOE's support for a grant to compensate for local and/or state sales and use tax actually paid by the EECF project or its members for tangible personal property that is a component of the project.

The exemption of F-T motor fuels and motor fuels blends containing F-T fuels from Section 4081 Federal Motor Fuel Excise Tax would give an incentive for the various states authorities to also eliminate state taxes. This combination of tax exemption would result in giving a margin advantage of the F-T fuel or fuel containing F-T blends of approximately \$.23 per gallon or \$9.66 per barrel. This would result in a projected increase of IRR on a commercial size EECF project producing 4,000 to 5,000 barrels per day of approximately 1.2%. This increased internal rate of return (IRR) may be the incentive to develop commercial size projects.

A DOE grant to compensate for local and/or state sales and use tax would result in increasing the IRR of the EECF project modeled in this section from 11.9% on the nominal basis to 12.3%.

Two of Texaco's technology licensors, Rentech and Syntroleum, have applied to the DOE to have F-T Synthesis products certified as "alternative fuels" under the Energy Policy Act (EPACT). The designation of fuel products containing at least 30% F-T synthesis products as "alternative fuels" under EPACT would help establish these fuels in the marketplace. The DOE's assistance in designating F-T diesel and blends containing F-T diesel as "alternative fuels" would be helpful in making the EECF a commercially feasible project.

This approach will allow the creation of small niche markets for F-T fuels in the order of magnitude of 500 to 2,000 barrels per day as F-T Synthesis plant capacity is increased over time. An example of a niche market would be the fuel supply for mass transit bus

system in a metropolitan area that may be in an EPA non-attainment area. The F-T blend diesel would produce lower emissions and allow the mass transit system to make the conversion to “alternative fuels” at a minimal cost since the same fueling infrastructure could be used and no engine conversion would be required.

## **Risk Assessment**

A risk assessment of the EECF facility was performed. The purpose of this effort was to test the interdependencies between units and the effect of individual and collective dependencies on the overall availability of the facility. The facility was modeled as a single train plant, with no spare equipment unless specifically indicated in the equipment list. Planned and unplanned failure data were solicited from team members for each of the elements modeled. Where data did not exist, industry sources were consulted to provide the reliability model with a reasonable basis. Overall planned downtime was not calculated on a process block-by-block basis, as any major outage in one block would have a direct effect on the availability of others. Planned downtime was calculated by identifying the longest planned outage for a given 6-month interval and identifying other planned events that could be performed during that outage, so that no additional downtime would be incurred for these shorter events. Those planned maintenance activities that did not fall into one of the 6-month turnaround intervals were counted as additional outages. Reliability prediction software was utilized to define the initial basis and generate reliability block flow diagrams. Data were inputted into this software for initial analysis and review, then loaded into Microsoft Excel for additional calculations, case comparisons, and development of charts and tables. The initial results indicate an overall EECF facility availability of 78% with an increase to 80.8% by sparing the Make-Up Hydrogen Compressor and Hydrogen Recycle Compressor in the F-T Product Upgrading Unit. Further improvements in overall availability may be obtained by selective sparing and/or parallel operation of some equipment. Additional work in this area will be conducted in Phase III.

## **Technical Barriers**

After the process design completed, an evaluation and assessment was performed to identify and assess the technical barriers that require additional research, development, and testing before the process can be commercialized. Several technical barriers have been identified within the F-T Synthesis Unit, Power Block, F-TPU, and AGR.

The barriers within the F-T Synthesis Unit have been categorized into four main areas: reactor design, catalyst/wax separation, equipment design, and environmental concerns.

Reactor design is critical to ensure that the synthesis reaction will produce the desired yields and offgas and liquid composition for the synthesis gas provided. The main technical barriers within this area are the confirmation of the catalyst performance, hydrodynamics, reactor scale-up, and the design of reactor internals. Catalyst performance has been included under reactor design because it is one of the most important considerations when designing a reactor. An acceptable reactor design will provide good distribution of the syngas within the reactor, manage the chemical energy that is released from the reaction, allow for the products to be removed from the reactor in a controlled fashion, and produce the maximum amount of desired products. The reactor should be the smallest that will maintain the desired reactor yield and selectivity.

Proper instrumentation must be provided in the reactor to measure slurry concentration, bed expansion, temperature gradients, gas holdup, etc.

Catalyst/wax separation is a critical issue that must be resolved. Currently, various separation methods are being evaluated outside of DOE EECF funding. For example, Texaco will demonstrate the effectiveness of the separation on a stand-alone system and a small slurry bubble column reactor (SBCR). Texaco will also privately fund construction and testing of a Demonstration Separator at the DOE Alternate Fuels Development Unit at LaPorte, Texas. Catalyst/wax separation involves the separation of the liquid products of the reaction from the catalyst. The purpose is to remove clean liquid products from the reactor while keeping the catalyst inventory within the reactor. The separation may occur within the reactor or may occur outside of the reactor. In the current design, the catalyst/wax separation is accomplished in two stages. The first stage removes the liquid products as filtrate while maintaining reactor catalyst inventory. The second stage removes the remaining catalyst solids from the liquid products being sent to the F-TPU. The design of the catalyst/wax separation system must be tested and optimized in phase II activities before the F-T Synthesis Unit can be commercialized.

Aside from the F-T Synthesis reactor and the catalyst/wax separation system, there are other equipment items within the F-T Synthesis Unit with unproven design. These items include the fresh catalyst activation vessel, the spent catalyst dump tank, and the catalyst withdrawal and replacement system.

In addition to the equipment design issues, there are some process design and environmental issues that need resolution. These include the disposal of F-T Synthesis catalyst and the use of F-T water in the gasification section. Testing will be required to confirm that the assumed methods are technically feasible.

The GE multi-shaft combined cycle STAG (steam and gas) 106FA system configuration designed for the EECF uses commercially available equipment except for the gas turbine combustion system. The GE 6FA gas turbine has been successfully tested for operation on syngas produced in Texaco gasifiers from a variety of feedstocks such as coal, petroleum coke, vacuum residue, etc. For the EECF proposed, part of the syngas from the Texaco Gasification Unit, after being treated in the AGR, will be sent to the F-T Synthesis Unit for conversion into liquid products. The tail gas from the F-T Synthesis Unit consists of uncondensed reactor products along with unreacted F-T feed gas. The thermal energy content of the F-T tail gas is lower than the syngas. There is no operating experience with the burning of this tail gas in the commercial combustion turbine. The percentage of F-T tail gas that can be burned along with the mixture of Texaco gasifier syngas and offgas from the AGR is not known. Moreover, the gas turbine combustor required in this design will also have air extraction for the ASU and nitrogen injection near the reaction zone for NO<sub>x</sub> emission control. The gas turbine combustion system will need to be designed and tested before it can be offered for commercial operation.

With regard to technical barriers for F-TPU, the Bechtel Wax Hy-Finishing™ technology has not yet been applied specifically to the hydrogenation of F-T liquids, although the concept of processing feeds of this composition is commercially proven. Summarily, the barriers to full confidence in the process technology to produce finished wax include: operating conditions of temperature, pressure, hydrogen to hydrocarbon ratio; deactivation rate of the process catalyst; separation of light hydrocarbon byproducts

produced during the process; and separation of the finished wax into narrow boiling and viscosity ranges without thermal degradation.

The primary technical risk with the AGR is the uncertainty of commercial tray efficiencies at medium pressure in stripping a rich amine with nitrogen at medium pressures. Error in this regard has a potential double negative impact that could result in less CO<sub>2</sub> to the gas turbine and more CO<sub>2</sub> to the SRU. Other potential risks include: trace or acidic contaminants in the syngas may cause amine degradation and/or corrosion problems, and trace oxygen in the stripping nitrogen may cause formation of heat stable salts and other degradation problems. These risks are judged relatively minimal since they could be addressed by design issues and/or available technologies that would mitigate or eliminate them.

In Phase I, solutions have been assumed for all of the above technical barriers. During Phase II, research, development, and testing will be performed to confirm and to develop solutions to the technical barriers that have been identified. Some of the work will be done within the scope of Phase II of the EECP and some will be performed outside the scope of the EECP.

## CONCLUSION

We believe the work to date looks promising. The goals to develop and commercialize coproduction technology that produces high value products, particularly those that are critical to our domestic fuel and power requirements, are progressing. And, resolution of critical knowledge and technology gaps on the integration of gasification and downstream processing are also moving forward. Further, there is high interest in the refining industry for this work because of its many synergistic benefits, such as use of a low-value feedstock, production of a variety of clean products, potential for lower overall plant emissions, and improved plant efficiency. In return, the EECP concept requires a facility such as a refinery to act as a heat sink to utilize heat (steam) produced in the process and to share infrastructure expenses. At this point in time, capital cost estimates for the facility are relatively high and work needs to focus on ways to reduce this hurdle. The economics for this small-scale facility are currently fairly marginal and thus, there may be a role for Government incentives to help foster further development of the concept to commercialization.

Site selection efforts culminated in the selection of the Port Arthur Refinery over the Tampa Electric Polk Station as the host site and the Finished Wax case as the optimum configuration. This selection was based on the most favorable economics over three other possible base cases and thirteen variations of those cases. The economics favored the Port Arthur Finished Wax case over the others primarily due to lower feedstock transportation costs, higher product value, and greater infrastructure compatibility. The conceptual design developed for this case during Phase I resulted in an overall cycle efficiency of approximately 66.5% and an early risk assessment predicts facility availability of 80.8%. Opportunities exist to increase both efficiency and availability further during Phases II and III.

A summary of the specific conclusions reached as a result of the process studies undertaken in Phase I follows:

- F-T Reactor Feed Gas Optimum H<sub>2</sub> to CO Ratio – Concluded that the additional expenditures required to increase the hydrogen content of the feed gas to the F-T reactor did not improve product yields or quantities sufficiently to improve overall plant economics.
- F-T Product Upgrading – F-T liquids can be considered premium feedstock with exceptional inherent qualities, such as negligible sulfur, nitrogen, and aromatics. After hydrocracking there is an increase in diesel production and a cetane number of more than 70. Niche market opportunities for this type of diesel may exist. Hydro-processing for wax production is suitable for this low volume, high value niche market product and helps the economics of this small-scale EECP facility.
- F-T Wastewater Treating/Utilization – It is believed that the F-T wastewater generated as a byproduct in the F-T reactor can be recycled to the Gasification Unit at low cost and negligible impact on the process or metallurgy. This will be confirmed in Phase II.
- F-T Catalyst/Wax Separation – There is not currently a proven commercial method for removing the catalyst from the wax. This is a critical issue that must be resolved. Currently, various separation methods are being evaluated outside DOE EECP funding. Prior to the detailed design and construction of the EECP Project, sufficient

work will have been identified for an effective means to separate catalyst and wax. Texaco will demonstrate the effectiveness of the separation on a stand-alone system, and a small SBCR outside of DOE funding. Texaco will also privately fund construction and testing of a Demonstration Separator for catalyst/wax separation at the DOE Alternate Fuels Development Unit at La Porte, Texas.

- F-T Reactor Feed Gas Percent CO<sub>2</sub> – It is expected that optimal F-T reactor performance is obtained with no CO<sub>2</sub> in the feed gas. Validation of this will be performed by laboratory experiments, and its effect or lack of effect will be confirmed in Phase II work.
- Process and Heat Integration Study – An energy assessment of the current plant configuration was made using pinch analysis. The analysis showed the design has a high degree of recoverable energy efficiency of about 92%. Opportunities exist to improve this further during Phase III work. Current process integration features include; recycle of F-T wastewater to the Gasification Unit, use of nitrogen from the Air Separation Unit (ASU) for gas turbine fuel dilution, use of nitrogen from the ASU as stripping gas in the Acid Gas Removal Unit (AGR), use of bleed air from the gas turbine combined cycle as feed to the ASU, and use of oxygen from the ASU in the Sulfur Recovery Unit (SRU). In addition to process stream integration, the following general heat integration steps were taken: low-grade heat from the gasifier was used to preheat boiler feed water for the heat recovery steam generator (HRSG); steam header levels in the process portion of the EECP were set to match HRSG, steam turbine, and host site header levels to the maximum possible extent; steam usage within the EECP was rationalized to minimize the number of levels at which the steam was exported or imported.
- F-T Tail Gas Utilization – Since electric power is a product of the EECP facility, F-T tail gas will be sent to the gas turbine and used for additional power production. In other situations the F-T tail gas could potentially be recycled to the gasifier, recycled to the F-T feed gas, or used as fuel in burners, depending on the process scheme and desired products, which can vary with the location.

The technical barriers discussed in Task 4.8 of this report and in the Research, Development, & Testing Plan, submitted under separate cover, have identified the key areas that need further investigation and development. The barriers associated with the F-T Synthesis Unit have been categorized into four main areas: reactor design, catalyst/wax separation, equipment design, and environmental concerns. Further, the fuel for the gas turbine is a combination of syngas, F-T tail gas, and AGR offgas, which has a lower thermal energy content than syngas alone. Thus, the gas turbine combustion system will need to be designed and tested before it can be offered for commercial operation. In addition, barriers to full confidence in the process technology to produce finished wax include: operating conditions of temperature, pressure, hydrogen to hydrocarbon ratio; deactivation rate of the process catalyst; separation of light hydrocarbon byproducts produced during the process; and separation of the finished wax into narrow boiling and viscosity range without thermal degradation. Solutions to these barriers will be addressed by the work covered under Phase II of the cooperative agreement.

The approach for the EECP was to develop a concept design that would have the greatest opportunity of approaching commercial criteria to ensure that the EECP would actually be built. The consortium of Texaco, Kellogg Brown Root, General Electric, Praxair, and Rentech agreed to design the EECP around a GE PG6101 (FA) gas turbine and a F-T reactor of 8- to 10-foot diameter that would resolve any reactor scale-up

issues for commercial applications. This design was based on a moderate-size electrical generation power unit for which there was reasonable data available to predict that the gas turbine combustor could handle the lower heating value fuel gas resulting from the integration of a power generation plant and a F-T synthesis process. The EECP concept study has confirmed that this approach is a reasonable path to reach the objective of a full-scale facility to demonstrate the EECP concept. For this size facility, we have shifted the F-T product mix to more valuable products to increase the revenue; however, the projected internal rate of return (IRR) of 11.9% will not meet any of the planned participants' economic criteria hurdles. A review of the Preliminary Project Financing Plan indicates that an IRR of at least 15.0 to 15.5% will be required to obtain interest from the financial community. From a lender's perspective, this IRR will allow the project to be leveraged in the 60% - 40% or 55% - 45% debt/equity range given the risk characteristics of the project.

The EECP as presently designed will require an additional non-equity contribution equivalent to \$70 MM of capital to reach the economic hurdle of 15.5% IRR. The project participants are interested in pursuing opportunities with the DOE to obtain financial support for the construction of the project. We recommend that discussions begin midway through Phase II after several critical milestones of the RD&T Plan have been reached.



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## LIST OF ACRONYMS

AFDU	Alternate Fuels Development Unit
AGR	Acid Gas Removal
AICHE	American Institute of Chemical Engineers
AP	Plant Air
API	American Petroleum Institute
ASU	Air Separation Unit
ATM	Atmosphere
BACT	Best Available Control Technology
bbl	Barrel
BD	Blowdown
BFD	Block Flow Diagram
BFW	Boiler Feed Water
bpd	Barrels per Day
bpsd	Barrels Per Stream Day
BRS	Brown & Root Services
BTM(s)	Bottom(s)
Btu	British Thermal Unit
BWF	Black Water Flash
CARB	California Air Resource Board
CBM	Coal Bed Methane
CCP	Commercial Coproduction Plant
CEQ	Council on Environmental Quality
CFCMS	Carbon Fiber Composite Molecular Sieve
CFD	Cubic Feet Per Day
CFI	Cold Flow Improvement Process
CFR	Code of Federal Regulation
CO <sub>2</sub>	Carbon Dioxide
COND	Condensate
CST	Condensing Steam Turbine
CT	Combustion Turbine
CW	Cooling Water
CWR	Cooling Water Return
CWS	Cooling Water Supply
DCFROI	Discounted Cash Flow Return on Investment
DEIS	Draft Environmental Impact Statement
DMPG	Dimethyl Ethers of Polyethylene Glycol
DOE	United States Department of Energy
EA	Environmental Assessment
EC	Economizer
EECP	Early Entrance Coproduction Plant
EIS	Environmental Impact Statement
EOR	Enhanced Oil Recovery
EOR	End of Run
EPA	Environmental Protection Agency
EPACT	Energy Policy Act
EPC	Engineering, Procurement and Construction
EPCM	Engineering, Procurement, Construction Management
ESD	Emergency Shutdown

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EV	Expected Value
EXCH	Exchanger
F	Degree in Fahrenheit
FCCU	Fluid Catalytic Cracking Unit
FDA	Food and Drug Administration
FDEP	Florida Department of Environmental Protection
FETC	Federal Energy Technology Center
FFBR	Fixed-Fluidized-Bed Reactor
FG	Fuel Gas
FONSI	Finding of No Significant Impact
FPCS	Functional Process Control Specifications
F-T	Fischer-Tropsch
F-TPU	Fischer-Tropsch Product Upgrade Unit
F-TR	Fischer-Tropsch Reactor
GE	General Electric Power Systems
GNHP	High Pressure Nitrogen
GNP	Gross National Product
GT	Gas Turbine
GTDL	Gas Turbine Development Laboratory
H <sub>2</sub>	Hydrogen
H <sub>2</sub> O	Water
H <sub>2</sub> S	Hydrogen Sulfide
H <sub>2</sub> SO <sub>4</sub>	Sulfuric Acid
HAP	Hazardous Air Pollutant
HC	Hydrocarbon
HCU	Hydrocracking Unit
HGU	Hydrogen Generation Unit
HHV	Higher Heating Value
HMX	Hybrid Multiphase Technology
HP	High Pressure
HRSG	Heat Recovery Steam Generator
HTR	Heater
HTU	Hydrotreating Unit
Hz	Hertz
IBP	Initial Boiling Point
ICC	Integrated Combined Cycle
IGCC	Integrated Gasification and Combined Cycle
IGV	Inlet Guide Vane
I/O	Input/Output
IP	Intermediate Pressure
IPCC	Intergovernmental Panel on Climate Change
IRCC	Integrated Reforming Combined Cycle
IRR	Internal Rate of Return
ISBL	Inside Battery Limits
JV	Joint Venture
K	Degree in Kelvin
KAAP	Kellogg Advanced Ammonia Process
KBR	Kellogg Brown & Root, Inc.
KBRTC	Kellogg Brown & Root Technology Center
KJ/hr	Kilojoule per hour
KO	Knock Out

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kPa	Kilo Pascal
KRES	Kellogg Reforming Exchanger System
KW	Kilowatts
LAB	Linear Alkyl Benzene
LAER	Lowest Achievable Emission Rate
LAO	Linear Alpha Olefins
LD	Liquidated Damage
LHSV	Liquid Hourly Space Velocity
LHV	Lower Heating Value
LNG	Liquefied Natural Gas
LP	Low Pressure
LPG	Liquefied Petroleum Gas
LS	Low Sulfur
LSFO	Low Sulfur Fuel Oil
LSVGO	Low Sulfur Vacuum Gas Oil
LTGC	Low Temperature Gas Cooling
M&EB	Mass and Energy Balances
MACT	Maximum Achievable Control Technology
MAPP	Mid-Continent Area Power Pool
MAT	Micro Activity Test
MDEA	Methyldiethanolamine
MEA	Monomethanolamine
mm	Millimeter
MM	Million(s)
MMBtu/hr	Million British Thermal Unit Per Hour
MMCFD	Million(s) Cubic Feet Per Day
MMKJ/hr	Million Kilojoules Per Hour
MMSCFD	Million(s) Standard Cubic Feet Per Day
MOC	Materials of Construction
MP	Medium Pressure
MSD	Material Selection Diagram
MT	Metric Ton
MTBE	Methyl Tertiary Butyl Ether
MTC	Montebello Technology Center
MT/D	Metric Tons Per Day
MTPD	Metric Tons Per Day
MW	Megawatts
MWE, MWe	Megawatts Electrical
N <sub>2</sub>	Nitrogen
NAAQS	National Ambient Air Quality Standard
NACE	National Association of Corrosion Engineers
NDT	Non-destructive Testing
NEPA	National Environmental Policy Act
NESHAP	National Emission Standards for Hazardous Air Pollutants
NG	Natural Gas
NH <sub>3</sub>	Ammonia
NIOSH	National Institute of Occupational Safety and Health
NLPM	Normal Liters Per Minute
Nm <sup>3</sup>	Normal Cubic Meter
NNF	Normally No Flow
NOC	Normal Operating Conditions

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NPDES	National Pollutant Discharge Elimination System
NPRA	National Petroleum Refiners Association
NPV	Net Present Value
NSPS	New Source Performance Standards
O <sub>2</sub>	Oxygen
O&M	Operations and Maintenance
OH	Overhead
OIS	Operator Interface Subsystem
OSBL	Outside Battery Limits
OSHA	Occupational Health and Safety Act
Pa	Pascal
PA	Pump-around
PAO	Polyalphaolefins
PAR	Port Arthur Refinery
PARFW	Port Arthur Refinery Finished Wax Case
PARHCU	Port Arthur Refinery Hydrocracking Case
PCF	Pounds Per Cubic Feet
PD	Positive Displacement
PDD	Project Design Data
PFD	Process Flow Diagram
PFS	Process Flow Sketch
PGDE	Polyethylene Glycol Dimethyl Ether
PGE	Power Generation Engineering
PI	Instrument Purge
PID	Piping and Instrument Diagram
PLC	Programmable Logic Controller
PLOT	Plot Plan
PM	Project Manager
PO/MTBE	Propylene Oxide/Methyl Tertiary Butyl Ether
PPMVD	Parts Per Millions Volume Dry
PPMVW	Parts Per Millions Volume Wet
PPMW	Parts Per Millions Weight
PSA	Pressure Swing Adsorption
PSD	Prevention of Significant Deterioration
psia	Pounds Per Square Inch Absolute
psig	Pounds Per Square Inch Gauge
PUC	Public Utilities Commission
R&D	Research and Development
RA	Relief to Atmosphere
RD&T	Research, Development and Testing
RFL	Liquid Refrigerant
RFV	Vapor Refrigerant
ROR	Rate of Return
ROSE <sup>®</sup>	Residual Oil Solvent Extraction
RVP	Reid Vapor Pressure
S50	50 psig Steam
S75	75 psig Steam
S150	150 psig Steam
S425	425 psig Steam
S900	900 psig Steam
S1500	1500 psig Steam

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SAP	Sulfuric Acid Plant
SBCR	Slurry Bubble Column Reactor
SCFD	Standard Cubic Feet Per Day
SCHP	High Pressure Steam Condensate
SCLP	Low Pressure Steam Condensate
SEM	Scanning Electron Microscopy
SEP	Separator
SH	Superheater
SH&E	Safety, Health & Environmental
SH600	Superheated 600 PSIG Steam
SIC	Standard Industrial Code
SIP	State Implementation Plan
SMR	Steam Methane Reforming
SOF	Soluble Organic Level
SPBC	Slurry Phase Bubble Column
SPP	Southwest Power Pool
SRU	Sulfur Recovery Unit
ST	Steam Turbine
STAG	Steam and Gas
STG	Storage
STM	Steam
STPD	Short Ton Per Day
STPR	Stripper
SWEPCO	Southwest Electric Power company
SWRI	Southwest Research Institute
SWS	Sour Water Stripper
T/D	Tons Per Day
TBA	Tertiary Butyl Alcohol
TBN	To Be Named
TBP	True Boiling Point
TDC	Texaco Development Company
TECO	Tampa Electric Company
TEMA	Tubular Exchanger Manufacturers Association
TESI	Texaco Energy Systems Inc.
TEWAC	Totally Enclosed Water/Air Cooled
TFBR	Tubular Fixed-Bed Reactor
TGFT	Texaco Gasification Fischer-Tropsch
TGP	Texaco Gasification Process
TGPS	Texaco Gasification Power System
TGTU	Tail Gas Treating Unit
THCU	Tampa Hydrocracking Case
TIC	Total Installed Cost
TNRCC	Texas Natural Resource Conservation Commission
TSC	Tampa Syncrude Case
TVA	Tennessee Valley Authority
UHC	Unburned Hydrocarbons
ULS	Ultra Low Sulfur
UOP	Universal Oil Products Inc.
U.S. EPA	U.S. Environmental Protection Agency
U.S. FDA	U.S. Food and Drug Administration
USGC	United States Gulf Coast

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UV	Ultraviolet
VI	Viscosity Index
VGO	Vacuum Gas Oil
VOC	Volatile Organic Compound
VPPM	Volume Parts Per Million
WBIP	Intermediate Pressure Boiler Feed Water
WBHP	High Pressure Boiler Feed Water
WBLP	Low Pressure Boiler Feed Water
WBMP	Medium Pressure Boiler Feed Water
WBS	Work Breakdown Structure
WGS	Water-Gas-Shift
WK	Clarified Water
WPPM	Weight Parts Per Million
WTI	West Texas Intermediate
WWTU	Wastewater Treating Unit
ZnO	Zinc Oxide